REQ. NO. 0-9143

SPOUTED BED GASIFICATION OF WESTERN CANADIAN COALS

CONTRACT REPORT NO. OSU81-00074

Contract Serial No. OSU81-00074
DSS File 20SU 23440-0-9143

Energy Research Laboratories, CANMET

FINAL REPORT 1981-82

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39pp

May 1982

ABSTRACT

This report describes the work done on CANMET contract OSU81-00074 "Spouted Bed Gasification of Western Canadian Coals" during the year 1981-82. Prior work was summarized in three previous Annual Reports. Gasification experiments in the 0.31-m dia. spouted bed were made to determine optimum operating conditions for different coals. To date successful operation has been confirmed with bituminous coals (Sukunka and Coleman), and sub-bituminous coals (Forestburg and Roselyn). Comparison of data gathered over the past three years gives a consistent picture of factors governing gasification of these coals. Sub-bituminous coal from the Klimax mine could not be handled due to sticking in the bed. This is attributed to formation of eutectics promoted by the high sodium content of the ash. The effects of recycle of cyclone char on the efficiency of gasification of Forestburg and Sukunka coals was investigated further. Performance of the gasifier in both spouting and fluidized bed modes is reported for three coals. A preliminary kinetics-free mathematical model of the gasification process is shown to give good predictions of gas quality. Some preliminary results of thermal fouling of the dust-laden gas coolers are presented. The report includes a description of modifications made to the gasifier unit, and the proposed experimental program for 1982-3.

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1. INTRODUCTION

This report describes the work done during 1981-82 on Contract OSU81-00074 of the Canadian Coal Conversion Programme, sponsored by CANMET. The main objective of this programme was to investigate the usefulness of a spouted bed gasifier for the treatment of Western Canadian coals. Specific objectives during 1981-82 included:

(i) Extend range of coals gasified to provide broader data bank on behaviour of Western Canadian coals

(ii) Compare spouted bed gasifier with fluidized bed gasifier

(iii) To improve the understanding of the process in the gasifier by obtaining gas composition profiles and temperature profiles within the reactor, and to develop a mathematical model to predict gasification performance.

(iv) Provide further data as required for gasifier scale-up.

The proposed work schedule is appended to this report. The details of the work done on the above mentioned items are described in this report.

2. PILOT PLANT MODIFICATIONS

A number of modifications were made on the gasifier pilot plant (Figure 1,2) during the year. Some of these were to improve the operation of the unit while others were done to achieve the specific objectives of the project.

2.1 Dust Recycling System

As reported earlier,[1-3] large dust carryover adversely affects the carbon conversion in the gasification process. This makes recycling of cyclone char vital. Lack of head room over the gasifier precludes mounting of the cyclones above the gasifier with a dip-leg arrangement for direct injection into the bed. A rotary valve originally constructed for top coal feeding of the gasifier was modified to permit recycling of the dust directly from the cyclone hopper to the gasifier. The changes were made essentially to minimize friction and accumulation of dust on the moving surface. This rotary valve typically operated for about half an hour before the fine coal dust trapped between the metal surfaces caused the valve to seize. Attempts to operate the rotary valve by feeding nitrogen as well as intermittent lubrication were also futile. It was concluded that we need a rotary valve with an outboard, dust tight, ball bearing for the dust recycle. Such a feeding system marketed by The Young Industries, Inc. of Muncy, Pennsylvania, is priced around US $4,000.

In the meantime an alternate system which could simulate the recycle of cyclone char was designed, fabricated and installed on the pilot plant. It consisted of a cylindrical hopper with conical bottom which was manually charged with dust from the cyclones. The dust from this hopper was fed to a vibra-feeder

and was then pneumatically conveyed to a reverse cyclone which fed it to the gasifier (Figure 3).

The dust inlet point in the gasifier was very close to the steam injection point. During the trial runs the dust from the simulated system could not be fed for more than a few minutes due to the condensed steam backing up in the feed line. To solve this problem, the dust recycle line was relocated and joined with the coal feed line. After this modification no problems were encountered in feeding the dust to the gasifier. However, at the vibra-feeder end it was difficult to control the dust feed rate. The two control variables i.e. (a) air pressure in the dust hopper (b) vibration setting on the feeder, had to be continuously manipulated during the run.

A table feeder (Figure 3) consisting of a slanted table which could be rotated at a desired speed by an external motor was then tried on the simulated dust recycle system. This feeder worked well and provided the dust feed rates within a reasonable limit. Most of the dust recycle runs were made using this system.

Although the simulated dust recycle system worked satisfactorily, it was not possible to directly relate the dust feed rates to the dust collection rates in such a system. For this purpose it is necessary to recycle the dust directly from the cyclone hopper to the gasifier. A special rotary valve has been ordered from The Young Industries, Pennsylvania, and will be installed soon.

2.2 Double Pipe Coolers

Two double pipe heat exchangers, similar to one already existing in the system, were designed, fabricated and installed (Figure 1) to replace the plate-coil tar condenser. The new heat exchangers improved the cooling of exit gas. Some attempts have been made to monitor the performance of these gas-coolers with time to get an idea of the fouling rates in such an application.

2.3 Refractory Lining of the Gasifier

The large number of experiments of short duration which had been run since the gasifier was commissioned inevitably resulted in cracking and chipping of the refractory lining. Patching has been done as required, however, a complete relining job became necessary. The lower cylindrical section of the gasifier and the cone were therefore relined with new refractory. These pieces were unbolted from the gasifier assembly, the old refractory was broken off and then the relining job was done by Plibrico (Canada) Ltd. The new refractory lining consists of two layers of 1/16" thick ceramic paper, 1" Duraboard insulating block and about 2.6" thick layer of Plibrico 80 S. This provides an inside diameter of 12" in the gasifier. The cone angle was fixed at 60°. The relining job was done in June 1981 and since then the new refractory has held up very well.

2.4 Fluidized Bed Operation

In order to operate the gasifier in fluidized bed mode a number of modifications were made. The design for fluidized operation was made in order to permit easy and rapid switching of the gasifier from spouted bed to fluidized bed operation.

A distributor was designed and fabricated using a 1.6 mm (1/16") thick titanium sheet. It contained 196 holes each 2.5 mm in diameter arranged on 2.0 cm x 1.73 cm triangular pitch. The distributor plate was placed between the conical and cylindrical sections of the gasifier. The details are shown in Figure 4. The coal feed line system has changed so that the fresh coal passed through the old solid sampling outlet of the spouted bed assembly. The coal feed entered the bed through a stainless steel pipe which terminated about 8 cm above the distributor. A cap was provided on the coal feed pipe to avoid blockage and minimize back flow of the material in this line when feeding is stopped. With this design it is possible to change the gasifier from spouted bed to fluidized bed operation and vice versa within 3 to 4 hours.

2.5 Fan-motor Assembly on Exhaust System

Operation of the off-gas incinerator has been a major problem. The system was re-piped in winter 1980, when deviations from the design drawings were found. On the suggestion of Lochhead-Haggerty, the designer and supplier of the incinerator assembly, a 3500 rpm motor was installed on the product gas exhaust system replacing the existing 1800 rpm motor. This was essentially done to increase the gas pressure going to the incinerator burner. Even after this change the pilot on the incinerator could not be kept lit.

The high speed motor was installed in May 1981. In November 1981 this motor failed. The bearings were found to be corroded and damaged, most likely by various chemicals and tarry materials present in the coal gas. Due to the delay in getting the 3500 rpm motor with adequate seals, a 1800 rpm motor was mounted. In January 1982, this motor burnt out due to water leakage in the body. At this stage the exhaust fan-motor assembly was re-built in our workshop and changed from direct drive to a belt drive system in order to protect the motor. In March 1982, the front bearing on the belt drive system was found to be corroded. It may be pointed out that our exhaust system is presently handling the flue gases from coal gasification pilot plant as well as from a much smaller experimental set-up on coal pyrolysis. The two failures of the fan-motor assembly in the recent months may have something to do with the more corrosive gases from the coal pyrolysis unit. A sealed bearing has been installed and an air circulation system is used to provide positive pressure around the bearing to minimize the entry of corrosive gases.

2.6 In Bed Gas Sampling

Provisional graduate student R. Srinivasan joined our group in January 1982. He has been designing and testing some gas sampling probes to allow radial and axial in-bed gas composition measurements to be made. His thesis will consist of measuring the profiles and comparing results with the mathematical models being derived with the help of Dr. Lim.

2.7 Future Modifications

The following changes will be made of the pilot plant to further improve its operability and to fulfill the objectives of 1982-83 programme.

(i) Install the rotary valve on the dust recycle system. Change the dust entry point from the high velocity spout to slow-moving annulus in the gasifier, if necessary.

(ii) Complete the testing and installation of pneumatic clamping system on the coal feeder.

(iii) Make necessary changes to permit the operation of pilot plant with oxygen and steam and produce a medium calorific value gas. This will involve construction of necessary facilities for a liquid oxygen tank, and complete piping for oxygen supply to the gasifier. As well the existing steam supply system to the gasifier must be enlarged.

(iv) Design, fabricate and install a jet scrubber between the heat exchangers and the drum filter.

(v) Improve the operation of the product gas incinerator by mounting a large incinerator burner in addition to the existing small pilot flame.

3. EXPERIMENTAL

The details of a typical gasification run in the spouted bed pilot plant have been described earlier (1). During fluidized bed gasification runs the basic procedure was the same as in the spouted bed experiments. 36.4 kg (80 lbs) of Ottawa Sand (bulk density = 1.56 gm/cc; particle density = 2.68 gm/cc) having an average particle diameter of 1.12 mm, was used an inert material during the fluidized bed experiments. It provided a fixed bed height of approximately 31 cm. The minimum fluidization velocity in air at room temperature was .63 m/s. The reason for using a smaller quantity of inert material during fluidized bed runs as compared to spouted bed runs (80 lbs vs 100 lbs of inert) was to maintain the same inert bed height in the cylindrical section of the gasifier. It must be realized that during fluidized bed runs the conical section lies below the distributor and there are no solids in this portion.

A series of runs was made under the fluidized bed mode with Forestburg and Sukunka coals. The gasification performance of these coals in the spouted bed has been reported earlier (2). Two new coals, a lignite from Saskatchewan (Klimax) and a sub-bituminous coal from Alberta (Roselyn) were also tested for their gasification performance in both spouted and fluidized bed gasifiers.

A few runs were made using dust recycle from the simulated system during gasification of Forestburg and Sukunka coals in the spouted bed mode. This provided data required to assess any improvements in the overall thermal efficiency of the gasification process by recycling the carryover char.

Table 3 summarizes all runs carried out during 1981-2.

4. RESULTS AND DISCUSSION

4.1 Recycle of Cyclone Char

In the operation of the gasifier under typical run conditions considerable carryover of fine carbon particles occurred. The freeboard velocity is typically .3 m/s, which exceeds the terminal velocity for particles smaller than 0.12 mm. The resulting carbon loss has an adverse effect on efficiency. A number of runs in which part of the dust was recycled to the gasifier using the simulated dust recycle system, demonstrated definite improvement in the hot gas thermal efficiency. These data using Forestburg and Sukunka coals are shown in Table 1. In all cases there is gain in thermal efficiency with char recycle, due primarily to increases in calorific value of the gas, and partly to increases in gas yield.

The experimental set-up for char recycling was described above in section 2.1. The present location of the dust entry point which is into the fresh coal feed line brings the recycled dust straight into the high velocity spout. This may result in re-entrainment of the recycled char with a resultant low residence time in the reactor. With the new recycle dust value system it is planned to re-locate the char inlet point to the gasifier annulus section where the solids are in down flow. This should increase the residence time of the char in the bed. The new system should permit direct recycle over sufficiently long operating times to demonstrate with more confidence the effect of char recycle on overall efficiency of the gasification process.

4.2 Gasification Throughput

In Figure 5a to 5d experimental data for Forestburg, Coleman, Sukunka and Roselyn coals have been plotted in terms of dry coal feed rate versus gas calorific value. These data cover experimental runs during 1979-80, 1980-81 and 1981-82. The air flow rate during these runs ranged between 119 - 137 kg/h. Some of the scatter is due to variations in air/coal ratio, particle size, steam/coal ratio etc. These results show an improvement in gas quality with increasing coal feed rate between the values 30 - 45 kg/h. When coal feed rate is increased beyond 45 kg/h the gas heating value decreases. These plots suggest that the optimum coal feed rate through 0.31 m dia spouted bed gasifier is about 45 kg/h on a dry basis irrespective of the coal type. For some coals e.g. Roselyn, feed rates above 45 kg/h were not run. The trend of improving gas quality as the coal rank decreases from the bituminous coals, Sukunka and Coleman, to the more reactive sub-bituminous Forestburg and Roselyn coals can also be seen from these figures. Further comments on this are made below.

Besides the coal feed rate, another prime operating variable during gasification is the air to coal ratio. This ratio, expressed on a dry coal basis, can be varied within a certain range during the gasification runs. For a given coal feed rate, the lower limit is dictated by the minimum air flow needed for maintaining the bed under spouting conditions. At high ratios of air to coal, gaseous H_2 and CO will be oxidized. The heating value of the gas produced from different coals is plotted as a function of air to coal ratio in Figures 6a to 6d. For all the four coals the gas heating value decreases with increasing air to coal ratio. Again these plots cover a range of less important variables, and include variations in feed rate, i.e. a given air/coal ratio may have been

run at coal feed rates of 30 or 40 kg/h. From the two sets of plots, the relative performance of the coals can be compared (Table 2). One can see that at a given Air/Coal ratio, e.g. 2.5 (Forestburg and Roselyn) the gas heating value of the lower rank coals is highest.

In addition to the calorific value of the gas produced in a given run, the dry gas yield is of considerable importance. The gas yield is almost uniquely determined by the Air/Coal ratio, and increases in a linear fashion with it, as shown in Figures 7a to 7c. The higher rank coal (Sukunka) gives a lower gas yield than does either Forestburg, or Roselyn, as might be expected.

4.3 Comparison of Spouted Bed vs Fluidized Bed Gasification

The gasification performances of Forestburg, Sukunka and Roselyn coals were assessed in both spouted and fluidized bed gasifiers.

The changes made in the pilot plant to operate it as a fluidized bed unit have been described in section 2.4 of this report, and the experimental conditions in section 3.0. The data for fluidized bed gasification of these coals are shown in Table 3. The fluidized bed can be operated with a fairly wide range of air to coal ratio (1.8 - 4.7). Data for fluidized versus spouted operation are also shown on Figures 5, 6 and 7. As with the spouted bed as the air to coal ratio is decrased, the gas heating value improves but the gas yield goes down. The overall effect is a slight decrease in the available energy per kilogram of dry coal. Examination of the above figures shows some unexpected results. For Sukunka coal (Figure 6c), the spouted bed appears slightly better overall in terms of gas quality than does the fluidized bed. For Forestburg coal (Figure 6a), at Air/Coal ratios above 2.5, there is virtually no difference in gas quality, and there is a suggestion that better results may be obtained in a spouted bed than a fluidized bed at Air/Coal < 2.5. For Roselyn coal (Figures 6d and 5d) the fluidized bed mode gave stronger gases than the spouted bed. For all three coals the differences between fluidized and spouted bed operation, in terms of gas heating value, are relatively small. Some further more detailed comparisons are made in Tables 4 and 5.

It may be concluded that both spouted and fluidized beds can successfully gasify caking and non-caking coals. The slight edge of one over the other depends upon the type of the coal. In general fluidized bed was found to give a better control on the gasifier operating conditions.

4.4 Gasification of Klimax Coal

Klimax coal was obtained from Manalta Coal Ltd. from Estevan, Saskatchewan. It ranks between lignite and sub-bituminous type coals and has a free swelling index of zero. Some of the properties of this coal are listed in Table 6.

Attempts were first made to gasify Klimax coal in the fluidized bed gasifier (Run No. 55). This coal showed a rather strange behaviour of inhibiting the bed movement which ultimately resulted in defluidization of the bed. In the second attempt the coal feed rate was kept at a very low level ($\simeq$ 10 kg/h) and the bed was carefully observed through the sight-glass. The air flow rate to the gasifier was kept at its normal settings for the fluidized bed operation. Within 15

minutes the frequency of the bed motion began to slow down and the solids appeared to become sticky rather than free flowing. De-fluidization quickly followed. The coal feed was stopped and the main air flow was increased from 30 to 60 cfm. Glowing agglomerates of 1 to 2 inches in size began to shoot up above the bed surface, bed movement increased gradually and the bed once more became fluidized. The coal feed was again started. However, within a few minutes the same damping behaviour was repeated. Even at a very low coal feed rate of about 10 kg/h and a fairly high air flow to the gasifier the bed could not be maintained in a fluidized state.

Based on the observation that after stopping the coal feed it was possible to re-fluidize the bed by feeding sufficient air, it was concluded that the agglomerates formed in the bed must be fairly loose. This conclusion was subsequently confirmed by inspecting the bed material after cooling to room temperature. It was found to contain loose agglomerates of 1 - 2 inches size which could easily be broken by hand. These were sent for chemical analysis, and the results are shown in Table 6. Klimax coal (4) contains fairly high quantities of sodium in its ash ($\simeq$ 8% Na_2O) as compared to other coals which have been gasified in our pilot plant ($\simeq$ 0.3 - 2.8% Na_2O). The sample of Klimax coal which we had received has not been analysed as yet for its ash constituents. It has been sent for 30 element semi-quantitative spectrographic analysis. Gasification of Klimax coal was subsequently attempted in the spouted bed. During these trials the same behaviour was noted as in the case of fluidized bed experiments and even the spouted bed could not be kept in motion. It was concluded that Klimax coal cannot be gasified in fluidized and spouted bed gasifiers under our normal operating conditions. Some changes in design of the gasifier such as addition of an agitator might lead to more satisfactory operation. The abnormal behaviour of this coal is attributed to the high sodium (alkali) content in its ash. From the Na_2O-SiO_2 phase diagram (5) it is apparent that a liquid phase can be formed at temperatures as low as 780°C. The liquid once formed can stick to the surface of inert particles and lead to the formation of agglomerates. Even in case of the three phase system of Na_2O-Al_2O_3-SiO_2 various liquid eutectics are formed starting at 732 C. Our interpretation is that when the coal feed is stopped and air is blown through the bed, fresh formation of liquid phase ceases and the liquid already present in the system changes to solid phase by gradual cooling of the bed. At this point the high air velocity in the bed is sufficient to break up the agglomerates and the bed starts fluidizing once again.

4.5 Gasification of Roselyn Coal

Roselyn coal was obtained from Manalta Coal Ltd. from Sheerness, Alberta. It is a sub-bituminous coal with a free swelling index of zero. Some of the properties of this coal are listed in Table 7.

Is is evident from the prior discussion in section 4.2 that gasification runs with Roselyn coal in both spouted and fluidized beds were successful. Referring to the plot of gas heating value vs air to coal ratio (Figure 6d), and to Table 2, gasification performance of Roselyn coal is seen to lie between Forestburg and Sukunka coals. A slightly better gas was produced in the fluidized bed gasifier as compared to the spouted bed. The fluidized bed also offered better control on the operating variables during the gasification run.

4.6 Performance of Gas Coolers

There is little available data on the performance of heat transfer equipment with dirty gases from coal gasifiers, although deterioration in heat flow with time can be a serious operating problem. Studies of the performance of the double pipe coolers downstream of the cyclone (Figure 1) were begun by Ken Stone, a chemical engineering final year student as his student thesis project. Temperature and flow rate data were gathered from a series of gasifier runs and the overall heat transfer coefficient between the solids-laden gas and cooling water determined as a function of time on stream. From such calculations the build-up of the thermal fouling resistance can be determined. Figure 8 shows a typical result covering several runs. The deposition of char particles on the heat transfer surface produces a thermal fouling resistance which grows nearly linearly with time. Cleaning of the exchanger restores its performance. Composite deposits accumulated over many gasifier runs were found to contain about 50% ash, 50% carbon at the hot end of the exchanger and about 30% ash, 70% carbon at the cold end of the exchanger. This work is being continued.

5. MATHEMATICAL MODEL OF THE SPOUTED BED GASIFIER

The aim of this work is to develop a detailed mathematical model for the spouted bed coal gasifier based on our experimental results. For a given gasifier size, coal, air and steam inputs, the final model should provide the temperature and composition profiles throughout the gasifier, as well as, the exit gas flow, composition and temperature.

A detailed model of this type will involve both reaction chemistry and gas and solids flow behaviour, and the success of the model will depend on the suitability of the parameters used. In order to assist in selection of parameters for the detailed model, we first considered two simpler models - a kinetics-free model, and a kinetic model which assumes the spouted bed is a well mixed reactor. To date, work on the kinetics-free model has been completed. The well mixed model has been formulated, however, only some preliminary results have been obtained, and these will not be discussed in this report. Work on the detailed model has been started.

Kinetics-Free Model

In this thermodynamic equilibrium model we assumed that the gasifier consists of two distinct zones, i.e. a gasification zone and devolatilation zone. The coal which enters the gasifier is rapidly heated and devolatilized. The remaining char and ash enter into the gasification zone. The devolatilization products bypass the gasification zone. The final exit gas compositions from the gasifier are found by adding the gasification products and devolatilization products.

5.1 Gasification Zone Products

In char gasification, we assumed that

a) All feed oxygen is consumed

b) No methane is formed by reaction of H and C

c) The Gases which emerge at the top of the gasification zone contain H_2, CO, CO_2, H_2O and N_2, and are in water-gas shift equilibrium

5.2 Devolatilization Products

We assumed that the fraction of the coal which is devolatilized is the same as the portion of the volatile matter determined by the proximate analysis.

Several methods were used initially to calculate the distribution of devolatilization products. The following method was found to provide the best agreement between experimental and calculated exit gas composition, especially with respect to methane concentration. Therefore it was used in the final calculation. This method of calculation is as follows:

a) The fractional distribution of chemical water and tar (including oil) are the same as those determined by the modified Fisher Assay of the coal used. The tar is assumed to have a chemical formula $C_1H_1O_{0.05}$

b) The remainder of the volatile matter is cracked to stoichiometric proportions of methane, carbon monoxide and hydrogen.

5.2 Model Inputs

Operating conditions:

1) Coal feed rate
2) Air/coal ratio
3) Steam/coal ratio

Experimental conditions:

1) Experimental measured temperature
2) Cyclone dust collection rate and carbon composition

Coal proximate analysis:

1) % volatiles
2) % fixed carbon

Coal ultimate analysis:

1) % carbon
2) % hydrogen
3) % oxygen

The gasification part of the kinetics-free model consists of a series of elemental balances for C, O, H and N, a statement of equilibrium for the water-gas shift reaction, and Daltons law. The 6 equations in 6 unknowns are solved by a non-linear equation routine on the UBC computer to yield molar flows and partial pressures.

5.4 Results and Discussion

Typical results for three different types of coal, and various experimental conditions are listed in Table 8. Agreement between experimental and calculated values is fairly good. Figures 9 (a-e) show experimental gas compositions plotted versus those calculated by the kinetics-free model. The agreement is seen to be fair, with measured results scattered about both sides of the 45° line. Only for H_2 is there a pattern of all experimental results being above the predicted values (by about 10%).

In the above results, the average bed temperature is used. Since there is a temperature profile in the spouted bed it was of interest to check the effect of temperature (which affects the equilibrium constant) on the predicted gas composition. A range of temperature values were assumed and results calculated for two runs are shown in Figure 10a and 10b. Both H_2 and CO_2 concentrations decreased with increasing assumed temperature whereas the CO concentration increases. The methane concentration which is dictated by the devolatilization process is unaffected. The variation of the exit gas composition of CO, CO_2 and H_2 is less than 1% per 100K change in assumed temperature. This indicates that it is safe to use the experimental average bed temperature for the model calculations.

6. PERSONNEL

The experimental work was carried out by Mr. Gordon Cheng, and Dr. C.B. Prakash, who was in charge of day to day operation. Dr. C. Jim Lim worked on the mathematical modeling on a part-time basis. Dr. A.P. Watkinson was the Principal Investigator.

7. PAPERS AND CONFERENCES

Dr. A.P. Watkinson presented a paper "Gasification of Coal in a Spouted Bed" at the I.G.T. Symposium "Advances in Coal Utilization Technology IV" held in Denver, April 1981. The paper will appear in the bound volumes from that symposium.

Dr. C.B. Prakash presented a paper "Gasification of Western Canadian Coals" at the District 6 Meeting of the C.I.M.M. in Victoria, October 1981. He also attended the Coal Technology '81 Conference and 4th Exposition of Coal Utilization Equipment in Houston in November 1981.

8. BUDGET

Table 9 is a budget summary for 1981-2. Budgeted amounts by category are compared with actual expenditures. We had to spend more on materials, supplies and equipment than expected, but made some savings on labour, subcontracts and travel in order to bring the project under the budgeted total.

9. FUTURE WORK

The work proposed for 1982-3 is included in a detailed work statement as Appendix 2.

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APPENDIX 1.

Work Statement 1981-82

Objectives

1. Extend range of coals gasified to provide broader data bank on behaviour of Western Canadian coals.
2. Compare spouted bed gasifier with fluidized bed gasifier.
3. Obtain in-bed gas composition and temperature profiles for use in mathematical model.
4. Extend mathematical model to predict gas composition as function of operating and design variables.
5. Provide further data as required for gasifier scale-up.

Major Tasks

1.	Design and test in-bed sampling rig for use in standard gasification tests.	1 Apr-1 Nov
2.	Revamp off-gas burner, dust recycle system and gas cooler. Design fluidization gas distributor.	1 Apr-31 May
3.	Install fluidized bed distributor and operate gasifier in fluidized mode with Forestburg and Sunkunka coals.	15 May-31 July
4.	Operate spouted bed gasifier with Crowsnest metallurgical coal feed.	15 Aug-15 Oct
5.	Operate spouted bed gasifier with Byron Creek thermal coal feed.	16 Oct-30 Nov
6.	Operate spouted bed gasifier with Alberta Roselyn coal.	1 Dec-31 Jan
7.	Operate spouted bed gasifier with Saskatchewan lignite.	1 Feb-31 Mar
8.	Run in-bed gas composition measurements.	1 Nov-31 Mar
9.	Improve gasifier mathematical model.	1 Apr-31 Mar
10.	Provide data as needed for scale-up.	1 Apr-31 Mar
11.	Prepare final report.	1 Mar-31 Mar

Spouted Bed Gasification of Western Canadian Coals

Work Statement 1982-3

Objectives

1. Extend range of coals gasified to provide broader data bank on behaviour of Western Canadian Coals.

2. Operate spouted bed on oxygen rather than air to produce a synthesis gas instead of a low calorific value gas.

3. Complete the modelling study of the spouted bed gasifier, and with it the experimental measurements of in-bed composition profiles.

4. Study the feasibility of sulphur removal by injection of limestone or dolomite into the gasifier.

5. Provide backup data and information as required for the design study of the 15 tonne/day gasifier.

Major Tasks

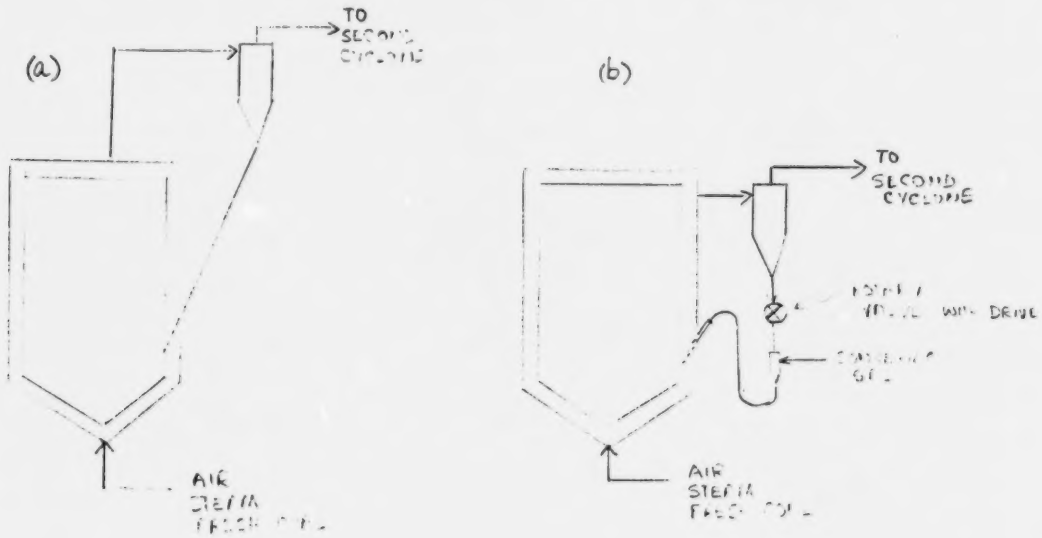
1. Four to six gasifier runs will be carried out with a bituminous coal from the Crowsnest coal field. The gasifier is a pilot scale operation and the experimental run frequency is a maximum of three runs in two weeks. The major tasks in carrying out the characterization of one coal are given below, and all the essential tasks must be repeated for characterization of other coals described in subsequent Major Task items. Sufficient coal received in lump form must be crushed and screened to yield about six 45-gallon drums of -6+18 mesh coal feed. This work is carried out by ourselves at the B.C. Research prior to trucking the coal to the gasifier site. Inert bed material must also be crushed and screened. The gasifier is prepared for a run by inspecting thermocouples, installing a new gas filter element, checking the chromatographs calibration, etc., charging the inerts and heating up the reactor with propane. The actual run takes place for a period of 4 to 6 hours, during which gas samples are taken, temperatures and pressures are recorded, and in some cases a series of programmed changes made to the operating conditions. At the end of the run the reactor is allowed to cool overnight. The bed is discharged the following day and the reactor cone removed, the in-bed char screened out, sieved to determine its size distribution, weighed and analysed for combustibles in a high temperature furnace. The coal feed drum is removed and contents weighed. The cyclone receivers are opened up, and the carry-over char weighed, sieved and a combustible determination done. Heat exchangers may be cleaned. Preliminary raw data calculations are done and all the flow rate, temperature, pressure and compositional data taken during the run (typically 12 - 16 determinations of gas composition,

.../2

for example,) are punched into the computer for processing. This includes temperature scans in the reactor and gas coolers. Computer output is analysed and plotted, and results of the run assessed. The preparations for the next run are then begun. To characterize a coal, a series of runs would be carried out at different air to coal ratios, and different coal feed rates.

1 April - 30 April

2. Short duration experiments have shown that a gain in efficiency is possible with recycle of product char from the first cyclone to the gasifier inlet. However, even with recycle the carbon content of the waste solids is high. The recycled char appears to blow rapidly through the gasifier if fed in through the high velocity air jet. Feed problems have been experienced. The low head room in the laboratory prevents the use of a direct cyclone/dip leg arrangement as shown in Figure (a). The set-up



shown in (b) will be attempted, in which a star-feeder will be used to meter dust from the cyclone outlet, and conveying gas will be used to introduce the feed into the annulus. This should result in a positive recycle to the gasifier, and the introduction of this feed into the annular region where it can readily react.

To make this modification involves the purchase of the rotary valve, its installation and calibration with char, the welding to the gasifier of a char inlet above the conical section, and the drilling of a suitable hole through the refractory lining, and connection of piping. Once installation and testing are complete a series of three or four runs is planned to determine whether injection of char into the annulus is better than injection into the spout with the incoming air, and to determine the extent of carbon conversion possible with char recycle over

.../3

extended runs of about 12 hours. A good ash balance which has not yet been accomplished with recycle, should be possible.
1 May - 15 June

3. The gasifier which now runs on air and steam producing a low calorific value gas, will be converted to permit operation with oxygen and steam and produce a medium calorific value gas. It is planned to have a dual connection system so that air can be used, for example during heat up, and then the oxygen flow switched on. Thus various combinations of oxygen and air can be used. In order to use essentially the same throughput of coal, and the same reactor lining, the flow of oxygen to the reactor will be the same as with air operation, but steam rather than nitrogen will moderate the temperature so the bed will not reach slagging temperature. The following changes will be made to the unit:

   i) The steam line will be enlarged and a new steam flow meter purchased and installed.
   ii) A fenced-off liquid oxygen tank platform will be constructed outside the Chemical Engineering building, after consultation with the oxygen supplier, and the University Physical Plant and Safety committee. Piping from the liquid oxygen tank will be run up to the laboratory.
   iii) A jet scrubber will be designed, fabricated and then installed between the heat exchangers and the particulate filter. A special water control system will be required to prevent occluded gases from being released into the laboratory water drain.

   Once these changes are made, and the new equipment tested, a series of runs will be made. 16 June - 31 July

4. The spouted bed gasifier will then be evaluated for production of medium calorific value gas. The first tests will be done with a non-caking coal such as Forestburg with which we have considerable experience with air-steam gasification. The effects of steam and oxygen to coal ratios will be of particular interest in these tests. About six runs will be carried out.
1 August - 30 Sept.

5. The system described in Major Task 3 will be used to treat a caking coal, for example, Sukunka in a repeat of Major Task 4. About six runs will be carried out. 1 Oct. - 30 Nov.

6. Tests of oxygen gasification of lignite will be carried out, as described in Major Task 1 and 4. 1 Dec. - 31 Jan.

.../4

7. Sulphur compounds in the coal are released during gasification primarily as H_2S and SO_2. These can be removed by wet-scrubbing process which treat cooled gas, or hot dry scrubbing processes. In another variant either limestone or dolomite can be injected into the bed to adsorb sulphurous compounds. Reactions are not as rapid as under oxidizing conditions however some sulphur removal is possible. In this task, a small feeder will be set up to inject limestone or dolomite particles into the bed to determine the extent of sulphur removal possible in the spouted bed. Local Texada Island limestone and dolomite from the B.C. interior will be tried. Preliminary runs will be restricted to one coal, and tests will probably be attempted first with air gasification. 1 Feb. - 31 March

8. During the course of Major Tasks 1, 2, and 4-7 an on-going program of radial and axial gas composition measurements will be made by graduate student R. Srinivasan. A variety of sampling probe designs are being considered during Fiscal Year 81-82, and some preliminary measurements will be made before F.Y. 1982-3 begins. The data from these measurements will be used as input into a mathematical models of the spouted bed gasifier now being developed. The compositional data will give a clearer picture of what is happening within the reactor, and will provide guidance to selecting the most appropriate model for scale-up calculations. Gases will be analysed for O_2, N_2, CO, CO_2, and CH_4. 1 April - 31 March

9. Current work has shown that a kinetics-free mathematical model using input data of coal feed rate, coal composition, air and steam rates and reactor temperature gives a good prediction of off-gas composition. This model can not predict carbon conversion, temperatures, effect of coal and blast feed rates etc. A kinetic model is being evolved which takes account of the gas flow in the spouted bed and the kinetics of the coal gasification reactions. The reactor is divided into a number of zones and mass and heat balances, fluid flow equations and kinetic expressions are applied. Further development work on the model will be carried out over the year to account for gas mixing and fluid dynamics in the bed and predict the temperature field in the reactor, the carbon conversion and the product distribution. 1 April - 31 March

10. Write a Final Report 1 March - 31 March

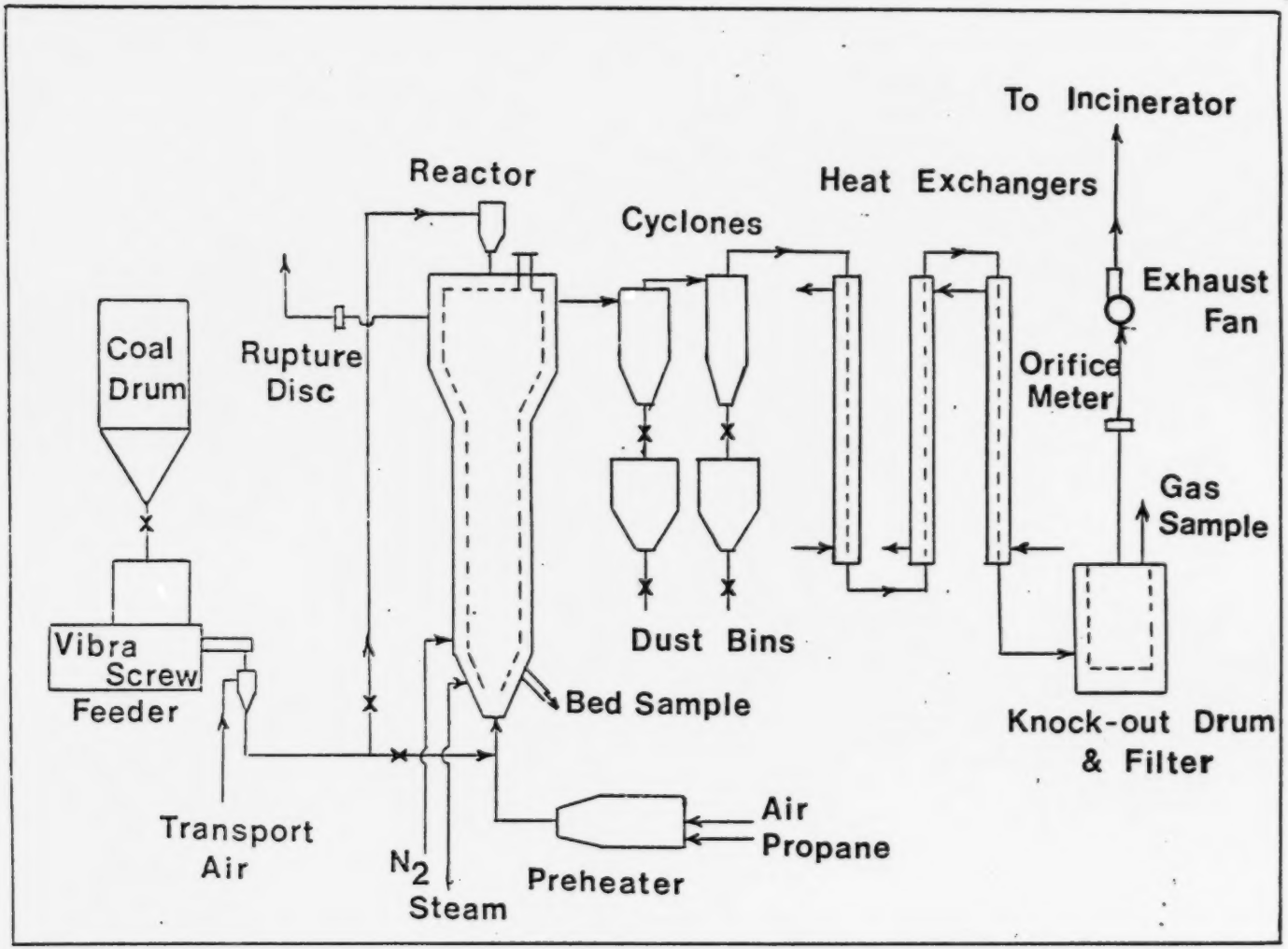


FIGURE 1

Flow Diagram of the Process Development Unit

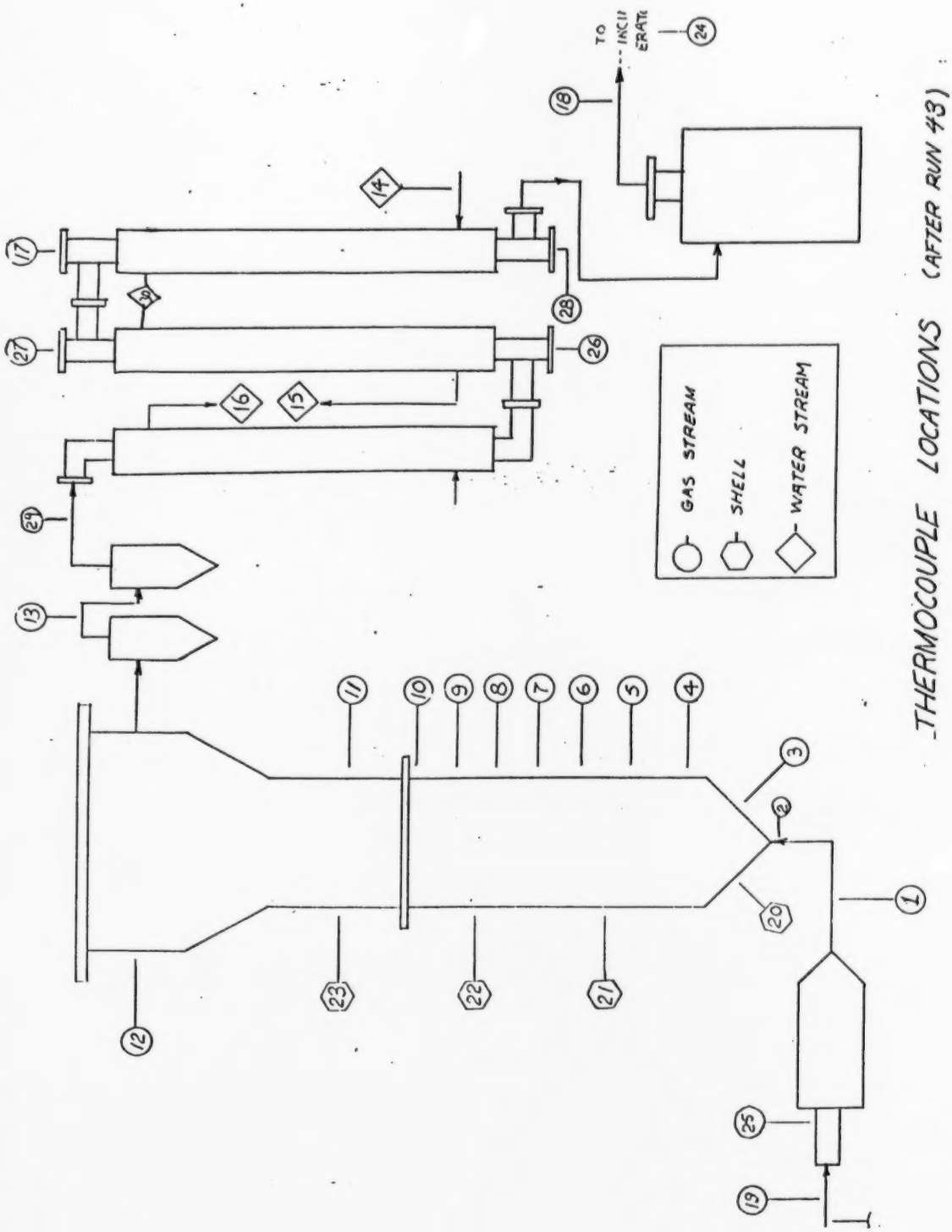
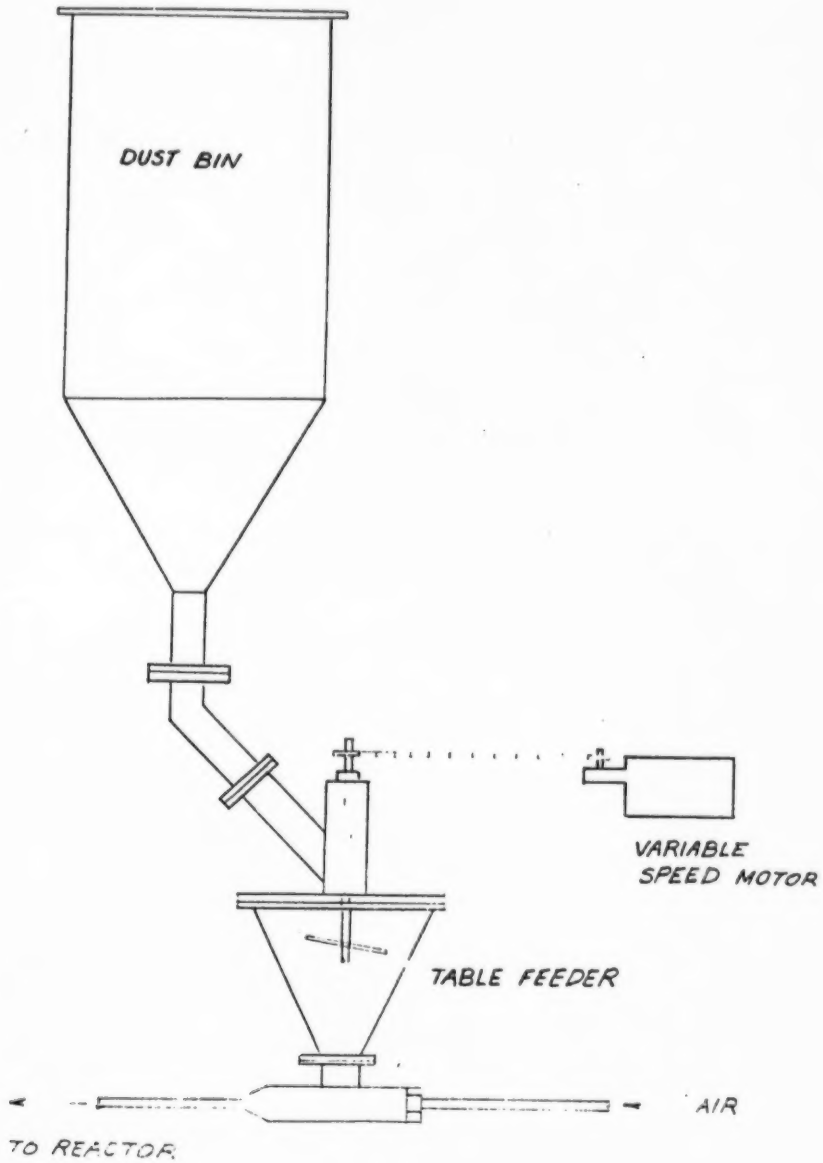


Figure 2. Thermocouple Location Diagram



SIMULATED DUST FEEDING SYSTEM

Figure 3. Cyclone Char Recycle Set Up

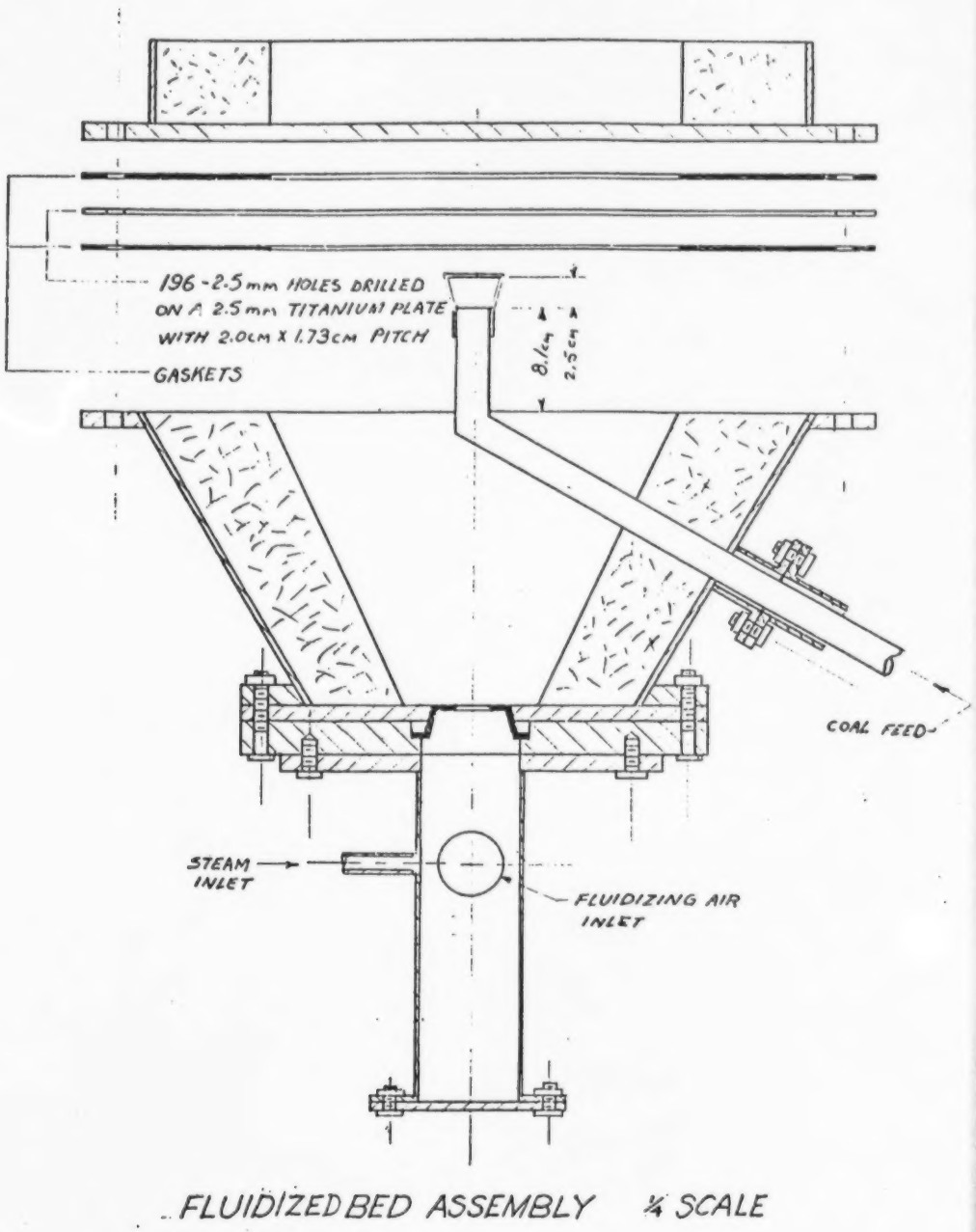
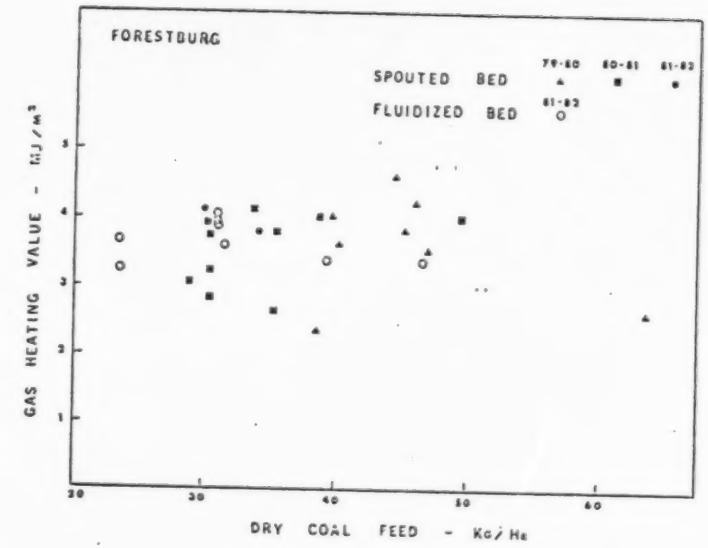
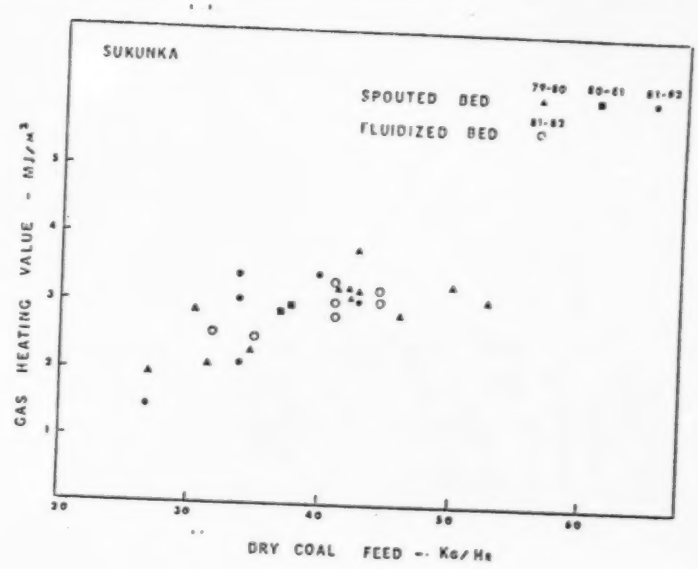
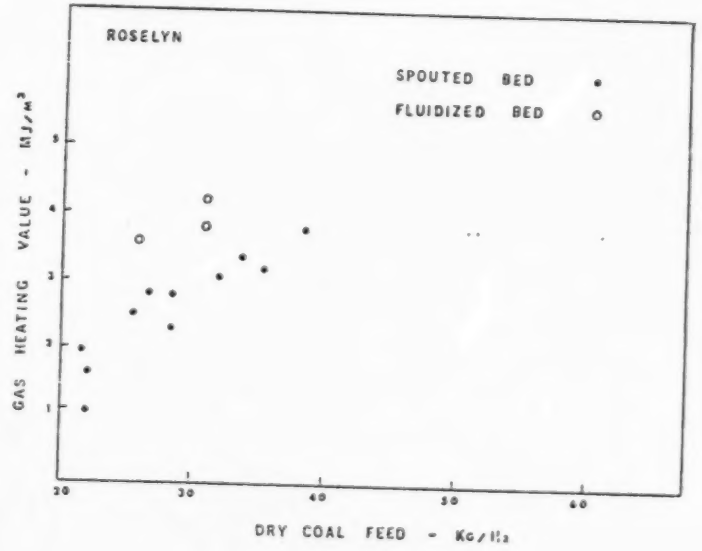


Figure 4. Distributor Plate Assembly for Fluidized Bed Operation







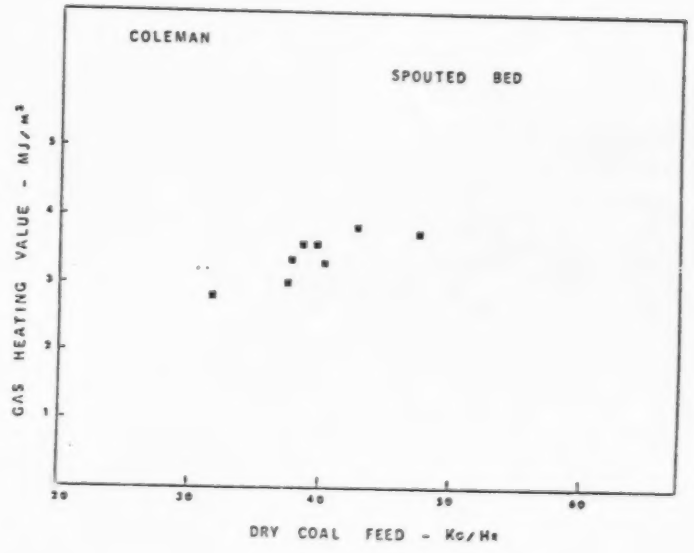


Figure 5 a-d. Dry Coal Feed Rate vs. Gas Heating Value Data for Forestburg, Coleman, Sukunka and Roselyn Coals.

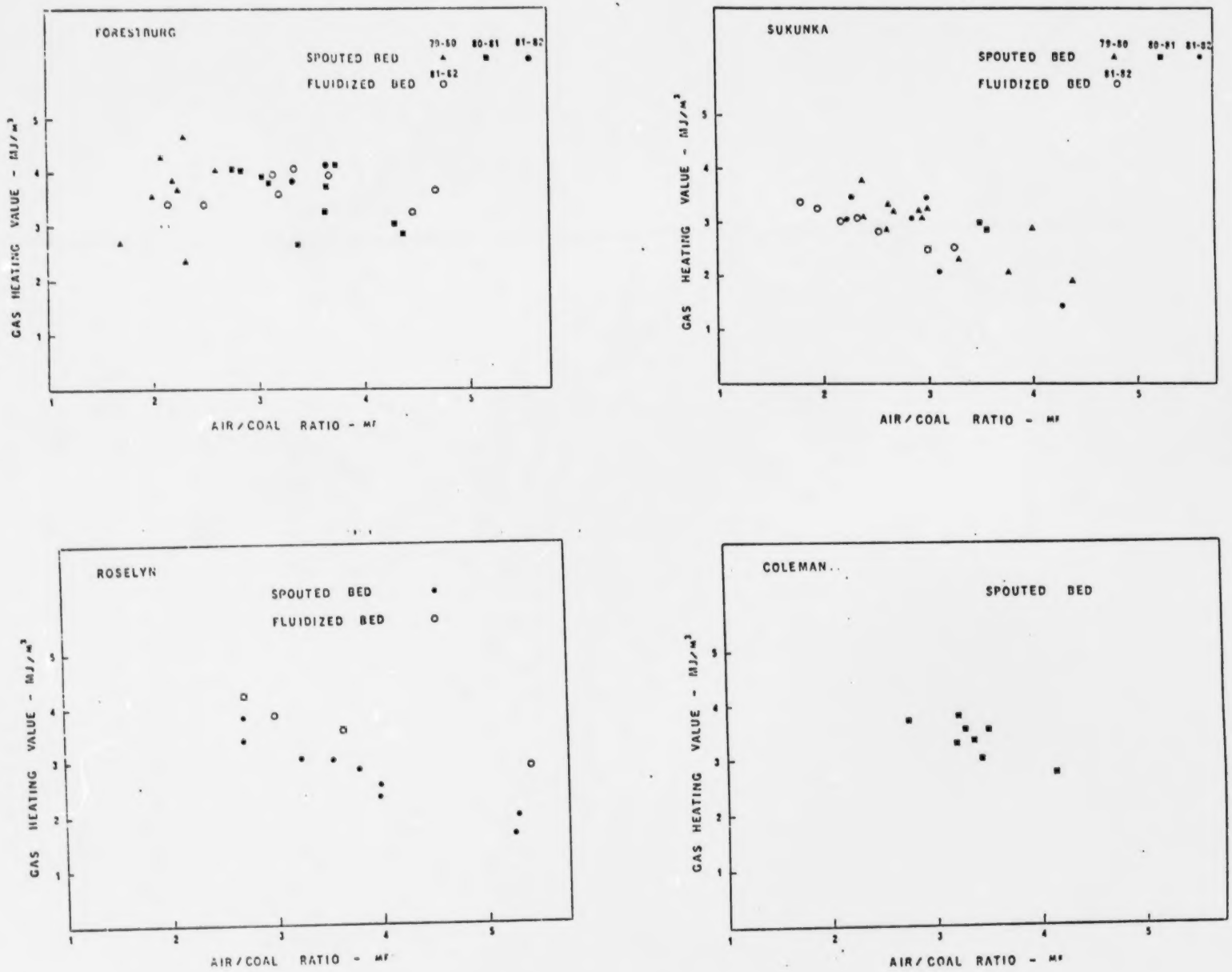
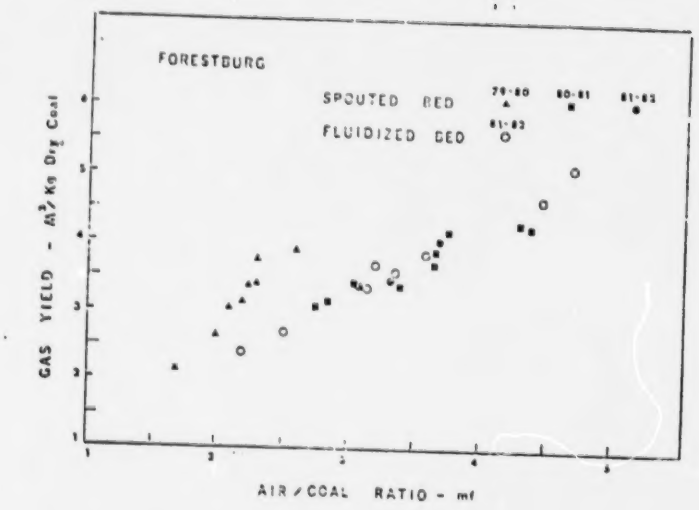
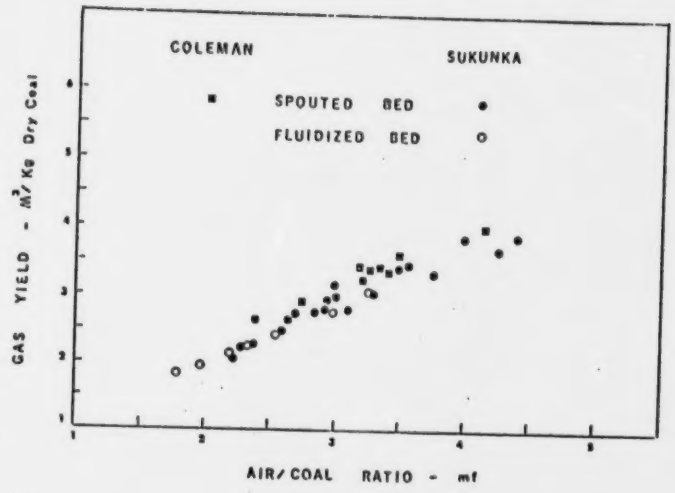


Figure 6 (a-d). Air/Coal Ratio vs Gas Heating Value Data for Forestburg, Coleman, Sukunka and Roselyn Coals.





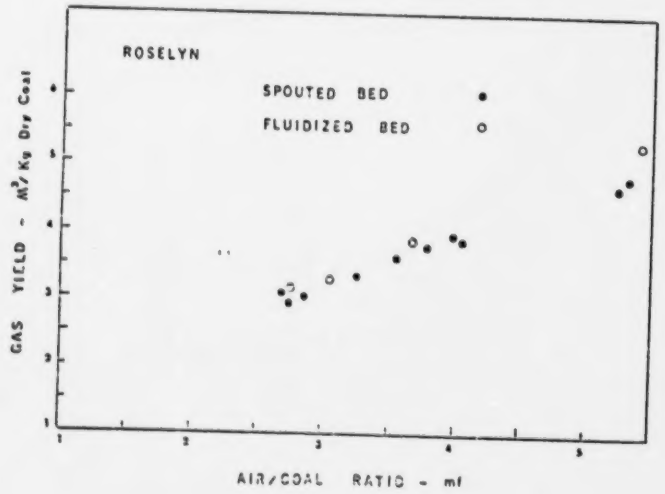


Figure 7. Air/Coal Ratios vs Gas Yield for a) Forestburg, b) Sukunka and c) Roselyn Coals

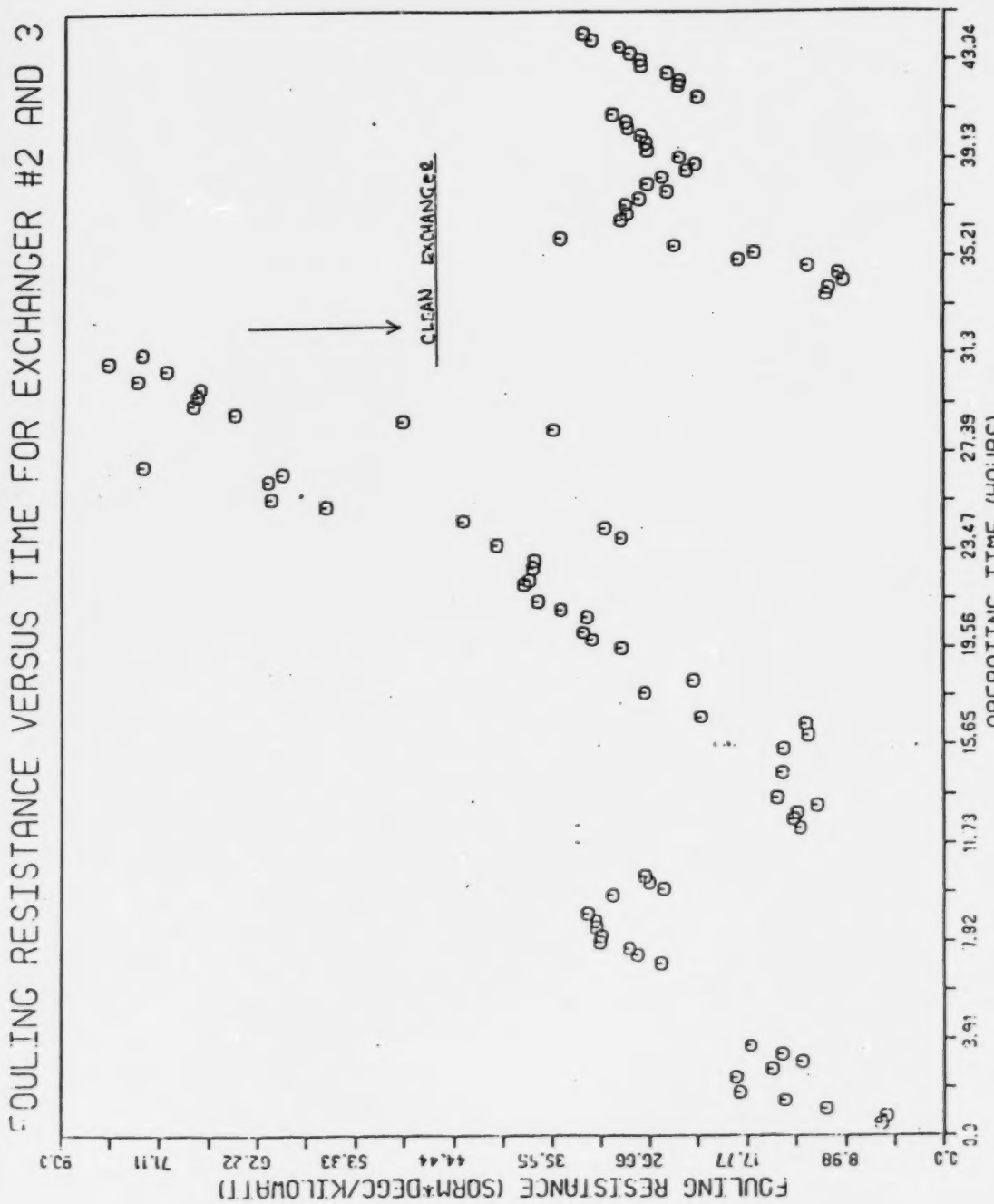
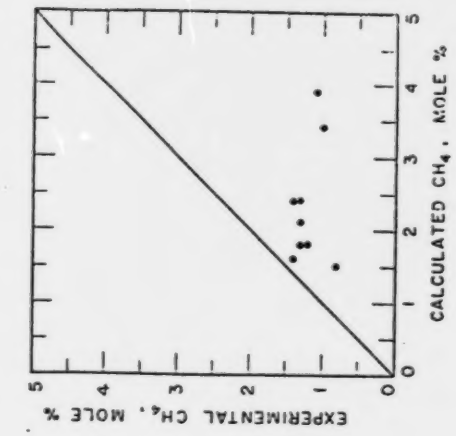
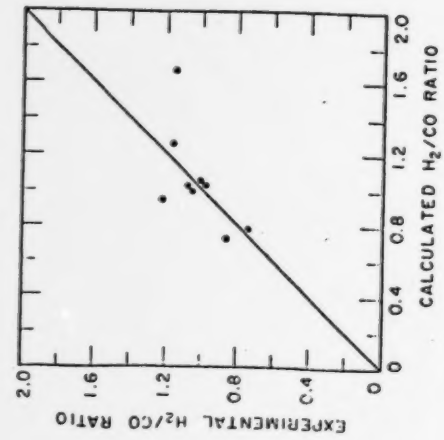
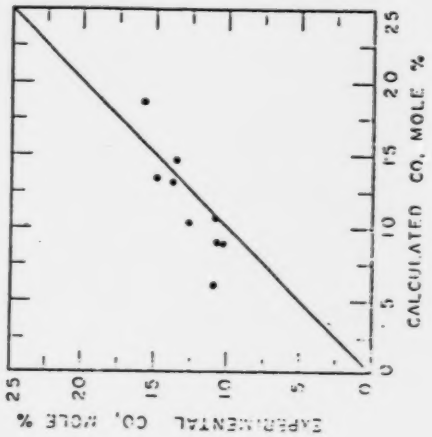
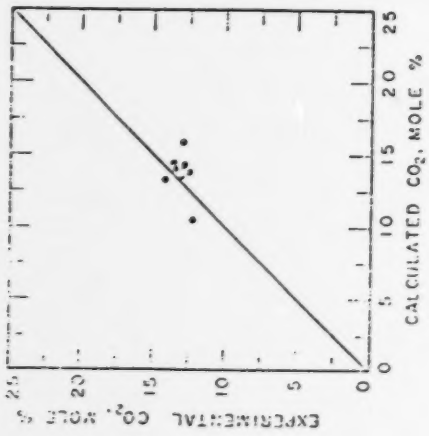


Figure 8. Fouling of Gas Coolers









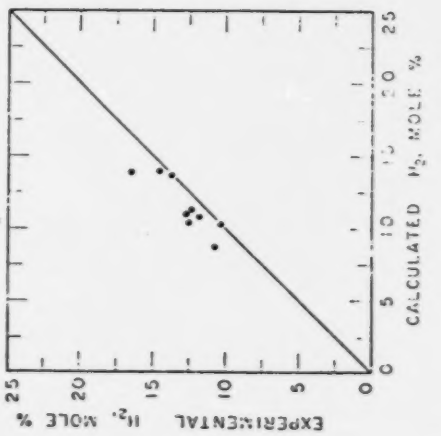
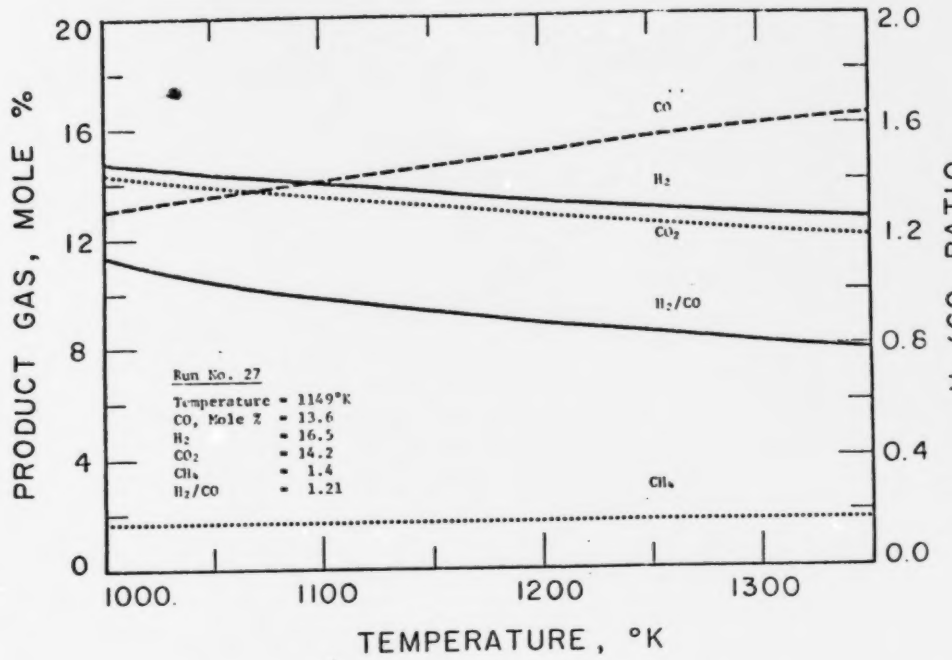


Figure 9. Experimental Gas Composition vs Predictions by the Kinetics-Free Model



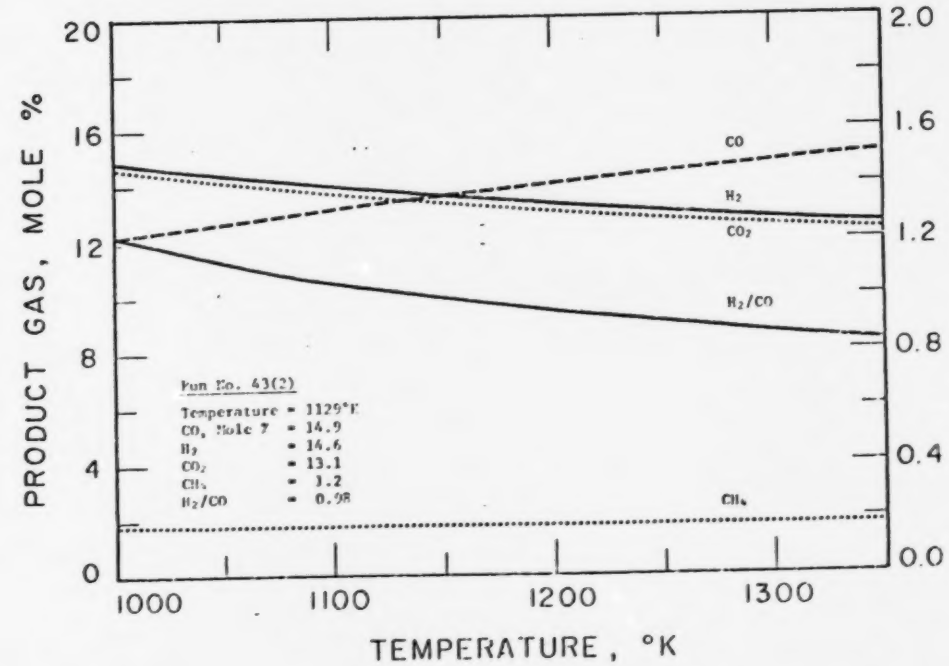


Figure 10. Effect of Assumed Temperature on Gas Composition

TABLE 1. EFFECT OF DUST RECYCLE

	Run	Dry Coal Feed Rate (kg/h)	Dust Recycle Rate (kg/h)	Gas Heating Value (MJ/m^3)	Gas Produced m^3/kg Dry Coal	Hot Gas Efficiency (%)
Forestburg	43(1)	35.2	0.0	3.81	3.40	68
	29	33.8	8.8	4.13	4.18	66
	47a	30.1	0.0	4.14	4.05	79
	47b	30.1	2.7	4.50	4.51	87
	48a	34.1	0.0	3.83	3.50	67
	48b	34.1	13.2	3.94	3.55	70
Sukunka	49a	26.7	0.0	1.46	3.59	35.4
	49b	26.7	1.02	1.78	3.76	42.3
	50a	33.8	0.0	2.09	2.75	31.9
	50b	33.8	20.9	3.06	2.99	43.2

TABLE 2. SUMMARY OF GAS HEATING VALUES*

Air/Coal Ratio	Coal			
	Sukunka	Coleman	Forestburg	Roselyn
2.5	3.5	3.75	4.1	3.9
3.0	3.3	3.5	4.0	3.5
3.5	3.0	3.4	3.8	3.0

* MJ/m^3 Dry Gas

TABLE 3 SUMMARY OF GASIFICATION DATA

Test No.	Duration of Run hr./min.	Gasifier Mode	Coal Type	Coal Size mm	% Ash in Coal %	Coal Feed Rate kg/h Wet	Coal Feed Rate kg/h Dry	Air/Coal Dry Basis	...r/Coal Dry Basis	Cyclone Catch kg/h	Cyclone Catch % Ash	Cyclone Catch %	Gas Yield m^3/kg Dry Basis	Gross Heating Value MJ/m^3	Thermal Efficiency Net Gas (%)	Avg. Bed Temp °C	Steam decomposed %	Product Gas Analysis Vol % Dry Basis H_2	CO	CH_4	CO_2	N_2	Comments
44	4:09	Sp. Bed	Sukunka	2.06	1.17	40.2	39.7	3.0	0.31	16.6	18.4	81.6	3.14	3.47	49.9	932	39.4	12.7	12.6	13.2	1.1	60.4	
45	2:10	Sp. Bed	Forestburg				Unsuccessful Dust Recycle																
46	2:00	Sp. Bed	Forestburg				Unsuccessful Dust Recycle																
47a	4:00	Sp. Bed	Forestburg	2.06	23.17	39.0	30.0	3.65	0.33	5.96	30.0	70.0	4.65	4.14	78.6	780	71.8	16.3	16.1	11.4	1.2	57.0	Avg. Bed Temp is obtained by taking average of thermocouples 7, 8, and 9
47b	Dust Recycle	Sp. Bed	Forestburg	2.06	23.17	39.0+2.68	30.0	3.90	0.30	5.96	30.0	70.0	4.51	4.50	86.6	829	79.8	16.6	18.6	10.9	1.2	54.7	
47c	Dust Recycle	Sp. Bed	Forestburg	2.06	23.17	29.7+2.68	22.8	4.95	0.47	5.94	30.0	70.0	5.02	3.39	83.0	916	54.8	11.2	12.6	12.9	1.4	61.9	
47d	Dust Recycle	Sp. Bed	Forestburg	2.06	23.17	33.0+4.08	25.4	4.41	0.49	12.05	30.0	70.0	4.63	3.67	82.2	946	53.2	12.4	14.3	11.6	1.1	60.6	
48a	4:30	Sp. Bed	Forestburg	2.06	23.17	44.4	34.1	3.33	0.30	5.96	35.0	65.0	3.50	3.83	66.7	777	52.9	12.5	16.0	10.1	1.0	60.4	
48b	Dust Recycle	Sp. Bed	Forestburg	2.06	23.17	44.4+13.16	34.1	3.34	0.30	13.19	35.0	65.0	3.55	3.94	69.9	818	54.8	12.8	17.7	9.4	0.7	59.4	
48c	Dust Recycle	Sp. Bed	Forestburg	2.06	23.17	38.7+1.49	29.7	3.84	0.30	13.19	35.0	65.0	3.85	3.34	69.1	877	44.1	10.0	16.5	10.2	0.4	62.9	
48d	Dust Recycle	Sp. Bed	Forestburg	2.06	23.17	38.7+1.49	29.7	3.86	0.63	13.19	35.0	65.0	3.77	2.86	63.4	891	32.9	11.0	10.9	13.1	0.6	64.4	
49a	4:10	Sp. Bed	Sukunka	2.06	1.08	27.0	26.7	4.27	0.53	7.50	28.9	71.1	3.59	1.46	35.4	905	17.7	6.3	4.1	13.9	0.5	75.2	Experienced great difficulty in feeding coal. Coal contains agglomerated fines which are causing problems.
49b	Dust Recycle	Sp. Bed	Sukunka	2.06	1.08	27.0+1.02	26.7	4.30	0.53	7.50	28.9	71.1	3.76	1.78	42.3	985	24.6	8.3	5.0	13.9	0.5	72.3	
50a	2:15	Sp. Bed	Sukunka	2.06	1.08	34.2	33.8	3.12	0.37	6.23	30.7	69.3	2.75	2.09	31.9	908	25.1	7.7	6.7	13.0	1.0	71.6	
50b	Dust Recycle	Sp. Bed	Sukunka	2.06	1.08	34.2+20.9	33.8	3.24	0.37	15.27	30.7	69.3	2.99	3.06	43.2	874	23.8	8.5	6.4	13.0	3.3	68.8	
51	1:25	Fl. Bed	Forestburg	2.06	23.17	30.60	23.51	4.47	0.35	3.61	42.2	57.8	4.47	3.26	78.6	774	54.8	10.5	14.7	13.2	0.6	61.0	
52a	4:00	Fl. Bed	Forestburg	2.06	23.17	30.60	23.51	4.70	0.50	4.20	46.4	53.6	5.13	3.68	95.0	852	58.5	12.5	14.5	13.8	1.1	58.1	Avg. bed temp is obtained by taking average of thermocouples 6, 7, 8, and 9
52b	4:00	Fl. Bed	Forestburg	2.06	23.17	40.20	30.89	3.58	0.30	4.20	46.4	53.6	3.88	3.96	75.1	887	54.3	11.8	17.7	11.2	1.1	58.4	
52c	4:00	Fl. Bed	Forestburg	2.06	23.17	40.20	30.89	3.34	0.30	4.20	46.4	53.6	3.60	4.06	68.3	889	51.5	11.9	18.4	10.2	1.1	58.4	
52d	4:00	Fl. Bed	Forestburg	2.06	23.17	40.20	30.89	3.15	0.30	4.20	46.4	53.6	3.38	3.95	63.2	923	45.3	11.1	18.6	9.9	0.9	59.5	
53a	3:33	Fl. Bed	Sukunka	2.06	1.08	35.40	35.02	3.00	0.18	11.91	30.7	69.3	2.78	2.50	34.6	922	31.6	8.6	7.7	13.4	1.4	68.9	
53b	3:33	Fl. Bed	Sukunka	2.06	1.08	45.00	44.51	2.35	0.15	11.91	30.7	69.3	2.27	3.08	33.0	934	38.3	11.1	9.3	12.6	1.6	65.4	
53c	3:33	Fl. Bed	Sukunka	2.06	1.08	45.00	44.51	1.98	0.15	11.91	30.7	69.3	1.94	3.23	29.4	929	35.1	11.9	8.8	12.6	1.9	64.8	
54a	4:15	Fl. Bed	Sukunka	2.06	1.08	32.10	31.75	3.27	0.33	18.38	34.2	65.8	3.07	2.56	39.5	906	32.3	9.6	7.8	13.7	1.3	67.7	
54b	4:15	Fl. Bed	Sukunka	2.06	1.08	41.40	40.95	2.56	0.26	18.38	34.2	65.8	2.44	2.84	33.7	903	33.2	11.1	8.3	13.3	1.3	66.0	
54c	4:15	Fl. Bed	Sukunka	2.06	1.08	41.40	40.95	2.19	0.26	18.38	34.2	65.8	2.15	3.01	31.1	901	31.8	12.1	8.7	13.0	1.3	64.9	
54d	4:15	Fl. Bed	Sukunka	2.06	1.08	41.40	40.95	1.81	0.26	18.38	34.2	65.8	1.84	3.35	27.2	734	31.4	13.9	9.4	12.8	1.4	62.5	Avg Bed Temp is low Pb got very low
55	-	Fl. Bed	Kilmax	2.06	35.86		Unsuccessful Run																
56a	3:00	Sp. Bed	Sukunka	2.06	1.08	43.2	42.73	2.22	0.11	14.77	16.9	83.1	2.14	3.03	30.2	886	34.4	10.0	8.8	12.7	2.0	66.5	
56b	3:00	Sp. Bed	Sukunka	2.06	1.08	34.2	33.83	2.84	0.17	14.77	16.9	83.1	2.77	3.08	39.3	904	44.1	11.1	10.0	12.3	1.4	65.2	
56c	3:00	Sp. Bed	Sukunka	2.06	1.08	34.2	33.83	2.29	0.17	14.77	16.9	83.1	2.31	3.44	35.2	880	43.1	13.1	10.2	12.2	1.7	62.8	
57a	5:00	Sp. Bed	Roselyn	2.10	27.01	29.4	21.46	5.32	0.87	2.82	50.0	50.0	4.85	1.94	64.0	962	24.0	7.4	7.3	15.7	0.4	69.2	
57b	5:00	Sp. Bed	Roselyn	2.10	27.01	36.6	26.71	3.80	0.76	2.82	50.0	50.0	3.78	2.83	61.4	944	32.5	11.7	9.3	14.9	0.8	63.3	
57c	5:00	Sp. Bed	Roselyn	2.10	27.01	46.2	33.72	2.74	0.73	2.82	50.0	50.0	2.93	3.37	53.3	904	34.1	15.3	8.8	16.0	1.2	58.7	
58a	3:10	Sp. Bed	Roselyn	2.10	27.01	30.0	21.9	5.94	0.95	3.70	36.8	63.2	5.0	1.03	48.3	897	11.2	3.6	3.8	16.9	0.3	75.4	
58b	3:10	Sp. Bed	Roselyn	2.10	27.01	30.0	21.9	5.25	0.95	3.70	36.8	63.2	4.7	1.63	56.3	917	18.1	6.1	5.7	16.6	0.5	71.0	
58c	3:10	Sp. Bed	Roselyn	2.10	27.01	39.0	28.47	4.05	0.81	3.70	36.8	63.2	3.85	2.31	54.7	895	24.7	9.1	7.8	15.6	0.7	66.7	
58d	3:10	Sp. Bed	Roselyn	2.10	27.01	39.0	28.47	3.55	0.61	3.70	36.8	63.2	3.39	2.6	56.9	874	29.4	11.6	8.8	16.2	0.9	62.5	
59a	3:45	Fl. Bed	Roselyn	2.10	27.01	24.0	17.52	5.43	0.71	2.79	41.2	58.8	5.37	2.93	51.8	840	32.9	14.2	9.4	15.9	0.9	59.6	
59b	3:45	Fl. Bed	Roselyn	2.10	27.01	33.4	25.84	3.68	0.37	2.79	41.2	58.8	3.91	3.58	83.2	864	42.4	10.2	11.1	14.0	0.8	63.9	
59c	3:45	Fl. Bed	Roselyn	2.10	27.01	42.6	31.09	3.05	0.42	2.79	41.2	58.8	3.30	3.83	69.5	861	43.2	12.2	13.8	13.5	1.2	59.3	
59d	3:45	Fl. Bed	Roselyn	2.10	27.01	42.5	30.98	2.74	0.37	2.79	41.2	58.8	3.16	4.22	61.3	841	44.9	12.2	15.6	12.4	1.3	58.5	
60a	3:40	Fl. Bed	Forestburg	2.06	23.17	40.8	31.35	3.21	0.40	5.56	33.2	66.8	3.72	3.01	60.2	708	54.1	14.3	16.7	12.6	1.3	55.1	
60b	3:40	Fl. Bed	Forestburg	2.06	23.17	51.0	39.18	2.50	0.39	5.56	33.2	66.8	2.70	3.41	69.2	790	48.4	13.4	13.9	16.6	0.8	55.3	
60c	3:40	Fl. Bed	Forestburg	2.06	23.17	60.6	46.56	2.16	0.37	5.56	33.2	66.8	2.40	3.39	68.3	745	35.7	13.1	11.2	15.4	1.3	59.0	
61a	3:18	Sp. Bed	Roselyn	2.10	27.01	34.8	25.6	4.03	0.89	5.14	26.6	73.4	3.94	2.34	43.3	750	32.8	13.5	11.0	16.8	1.1	57.6	
61b	3:18	Sp. Bed	Roselyn	2.10	27.01	43.8	31.97	3.24	0.78	5.14	26.6	73.4	3.37	3.09	57.7	863	27.7	10.7	7.5	16.0	0.9	64.9	
61c	3:18	Sp. Bed	Roselyn	2.10	27.01	52.8	38.54	2.69	0.67	5.14	26.6	73.4	2.81	3.85	54.9	809	31.0	12.7	9.4	15.8	1.1	61.0	
															58.2	778	43.7	17.2	10.8	14.2	1.2	54.6	

TABLE 4 COMPARISON OF SPOUTED AND FLUIDIZED BEDS

	FOREST BURG		FOREST BURG	
	SPOUTED BED	FLUIDIZED BED	SPOUTED BED	FLUIDIZED BED
RUN NO.	42(2)	52	42(3)	52
DRY COAL FEEDRATE kg/h	30.5	30.9	30.5	30.9
AIR/COAL (Dry basis)	3.66	3.58	3.05	3.15
STEAM/COAL (Dry basis)	0.45	0.30	0.50	0.30
AVE. BED TEMP. °C	919	887	870	920
GAS YIELD m^3/kg dry coal	3.92	3.88	3.39	3.38
HEATING VALUE MJ/m^3	3.73	3.96	3.93	3.94
CYCLONE CATCH kg/kg coal	0.110	0.137	0.110	0.137
COLD GAS EFFICIENCY %	56.2	58.0	50.5	50.9
STEAM DECOMPOSED %	48.7	54.5	44.7	45.3
SUP. VEL. AT STD. CONDITIONS m/s	0.35	0.33	0.30	0.29
GAS COMPOSITION % VOL.				
H_2	12.3	11.6	15.3	11.1
CO	15.2	17.7	13.6	18.6
CO_2	11.3	11.2	12.9	9.9
CH_4	1.0	1.1	1.1	0.9
N_2	60.2	58.4	57.1	59.5
H_2/CO RATIO	0.81	0.66	1.10	0.6

TABLE 5 COMPARISON OF SPOUTED AND FLUIDIZED BEDS

	SUKUNKA		SUKUNKA	
	SPOUTED BED	FLUIDIZED BED	SPOUTED BED	FLUIDIZED BED
RUN NO.	30	53	15	54
DRY COAL FEEDRATE kg/h	37.7	35.0	41.1	41.0
AIR/COAL (Dry basis)	3.5	3.0	2.96	2.56
STEAM/COAL (Dry basis)	0.33	0.18	0.26	0.26
AVE. BED TEMP. °C	965	922	969	903
GAS YIELD m^3/kg dry coal	3.4	2.78	3.0	2.44
HEATING VALUE MJ/m^3	3.0	2.50	3.28	2.84
CYCLONE CATCH kg/kg coal	0.15	0.34	0.31	0.45
COLD GAS EFFICIENCY %	32.1	21.9	30.8	21.8
STEAM DECOMPOSED %	37.8	31.6	44.4	33.2
SUP. VEL. AT STD. CONDITIONS m/s	0.44	0.34	0.41	0.36
GAS COMPOSITION % VOL.				
H_2	11.0	8.6	11.4	11.1
CO	9.1	7.7	8.2	8.3
CO_2	13.4	13.4	11.2	13.3
CH_4	1.5	1.4	1.4	1.2
N_2	65.0	68.9	67.8	67.7
H_2/CO RATIO	1.2	1.1	1.4	1.3

TABLE 6. PROPERTIES OF KLIMAX COAL

A. Proximate Analysis

	Moisture %	Ash %	Volatile Matter %	Fixed C %	Calorific Value BTU/lb
As received	35.86	5.23	25.19	33.72	7390
Air dry	12.13	7.16	34.51	46.20	10124
Dry	-	8.15	39.27	52.58	11520

B. Ultimate Analysis

	Moisture %	Ash %	C %	H %	N %	S %	O %
Air dry	12.13	7.15	60.66	4.83	1.06	0.66	25.64
Dry	-	8.14	69.03	3.95	1.21	0.75	16.92

C. Free Swelling Index: 0

D. Ash Fusion (°F in reducing atmosphere)

IT	2082
ST	2137
HT	2162
FT	2475

E. Analysis of agglomerate
Na_2O = 0.7; K_2O = 0.04; P_2O_5 = 0.41; SiO_2 = 56.0
The balance may be other oxides like Al_2O_3, CaO, MgO etc.

TABLE 7. PROPERTIES OF ROSELYN COAL

A. Proximate Analysis

	Moisture %	Ash %	Volatile Matter %	Fixed C %	Calorific Value BTU/lb
As received	27.01	3.98	29.11	39.90	8714
Air dry	10.71	4.87	35.61	48.81	10660
Dry	-	5.45	39.88	54.67	11939

B. Ultimate Analysis

	Moisture %	Ash %	C %	H %	N %	S %	O %
Air dry	10.71	4.87	63.70	5.16	1.28	0.62	24.37
Dry	-	5.45	71.34	4.44	1.43	0.69	16.65

C. Free Swelling Index: 0

D. Ash Fusion (°F in reducing atmosphere)

IT	2060
ST	2090
HT	2130
FT	2355

TABLE 8. COMPARISON OF EXPERIMENTAL AND KINETICS-FREE MODEL RESULTS

	FORESTBURG				COLEMAN	
	RUN 28		RUN 43(2)		RUN 31	
	CALC.	EXPT'L	CALC.	EXPTL	CALC.	EXPTL
CO (% Vol)	18.6	15.8	13.4	14.9	9.0	10.7
H_2 "	13.6	13.7	13.8	14.6	11.3	12.4
CO_2 "	10.6	12.2	13.7	13.1	14.3	12.9
CH_4 "	1.8	1.3	1.8	1.2	2.4	1.4
N_2 "	55.5	57.0	57.2	56.2	63.0	62.6
GAS YIELD m^3/kg coal	3.07	3.09	3.05	3.18	3.12	3.24
HEATING VAL. MJ/m^3	4.56	4.04	3.98	4.01	3.37	3.31
H_2 / CO	0.732	0.867	1.03	0.98	1.26	1.16

	COLEMAN		SUKUNKA			
	RUN 33		RUN 34		RUN 44	
	CALC.	EXPTL.	CALC.	EXPTL.	CALC.	EXPTL.
CO (% Vol)	10.6	10.8	8.9	10.2	10.3	12.6
H_2	10.8	11.7	8.8	10.7	11.0	12.7
CO_2	13.8	13.3	14.3	13.5	13.5	13.2
CH_4	2.1	1.2	3.4	1.0	3.9	1.1
N_2	62.7	63.0	64.6	64.6	61.3	60.4
GAS YIELD m^3/kg coal	3.70	3.81	3.36	3.44	3.00	3.14
HEATING VALUE MJ/m^3	3.36	3.16	3.43	2.90	4.02	3.46
H_2 / CO	1.02	1.08	0.99	1.05	1.07	1.01

Table 9. Budget Summary 1981-2

		Budgeted $	Spent* $
A.	Labour (Including Benefits)		
	Research Associate	27,720	26,400
	Research Assistant	22,995	21,900
	Graduate Student (+ Part time Assoc)	8,400	6,493
	Principal Investigator	4,029	4,029
	Benefits		2,418.77
	Subtotal	63,144	61,241.14
B.	Materials and Supplies	2,300	4,598.58
C.	Equipment (Includes repair charges)	4,800	5,847.41
D.	Computer	500	95.00
E.	Subcontract (Coal Analysis, Preparation & Shipping)	5,500	4,021.64
F.	Travel and Living	3,000	2,429.78
G.	Overheads (30% of A + B, 2% of F)	19,690	19,800.51
	Total	98,937	98,034.06

* This summary is unofficial, and subject to correction on the Final Accounting statements from U.B.C. Department of Finance.

POOR

QUALITY

ORIGINAL